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(54) Title: ABSORPTION PROCESS WITHOUT EXTERNAL SOLVENT		
(57) Abstract <p>An absorption process for separating a feed gas stream having components with a spectrum of volatilities including light (volatile) components, intermediate volatility components, and heavy (least volatile) components. The disclosed process includes the steps of: (1) contacting the feed gas stream with an absorption solvent stream in an absorber to produce a light product gas stream that contains predominantly light components, and a rich solvent stream containing intermediate and heavy components absorbed from the feed in the absorption zone; (2) flashing the rich solvent stream to reduced pressure in a flash zone to produce a flash gas comprised of intermediate and heavy components, and a liquid stream consisting essentially of heavy components; and (3) conveying the liquid stream from the flashing zone to the absorber as the absorption solvent; wherein the absorption solvent is comprised of heavy components taken from the feed; the process does not use an external absorption solvent. The process is particularly useful for rejecting nitrogen from natural gas and for recovering hydrogen from refinery and petrochemical process off-gases.</p>		

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10 **ABSORPTION PROCESS WITHOUT EXTERNAL SOLVENT**

 This invention relates to removing nitrogen from natural gas and to
recovering hydrogen from refinery and petrochemical plant off-gas streams using
15 physical absorption processes.

Nitrogen Rejection from Natural Gas

 Natural gas is a mixture of methane and light hydrocarbons (ethane,
propane, butanes, pentanes, and small amounts of hexanes, heptanes and heavier
20 components). Nitrogen is commonly found in natural gas, often in significant
concentration. Other components of natural gas include helium, hydrogen sulfide,
carbon dioxide, and water. Occasionally, cyclic and aromatic compounds occur in
natural gas.

25 The composition of natural gas streams vary over a wide range as indicated
by the data presented in Table 1. Nitrogen content in natural gas varies from a
few percent to over 50%. The nitrogen may have formed naturally or have been
introduced by injection into oil wells to stimulate production.

30 Gas pipeline transmission systems limit the allowable maximum content of
non-combustible gases, nitrogen and carbon dioxide in natural gas because non-
combustible gases consume pipeline capacity and transmission power but have no
heating value. The maximum specification for non-combustibles ranges typically
between about 2 mol % and about 8 mol %. The Gas Research Institute estimates

that there are about 55 trillion cubic feet of natural gas reserves in the continental United States that are "sub-quality". The Gas Research Institute defines natural gas containing more than 4 mole % nitrogen as sub-quality. These gas reserves must be processed to reject nitrogen and other non-combustibles to be acceptable to most pipelines. In addition, production of oil that is associated with high nitrogen gas is impeded because government regulations bar producers from burning the nitrogen-rich gas or venting the gas into the atmosphere.

Currently, the predominant technology for rejecting nitrogen from natural gas is cryogenic. The natural gas feed is liquefied by flashing the gas across Joule-Thompson valves or expanders to achieve very low temperatures (as low as -185°C (-300°F)). Then, nitrogen is fractionated overhead out of the liquefied natural gas in a distillation tower forming a bottoms stream of methane with low nitrogen content. Cryogenic processes are intensively heat exchange integrated to reduce power consumption, e.g., incoming feed is cooled by heat exchange against outgoing cold product streams, thus conserving most of the energy that went into refrigerating the product.

Nitrogen rejection adds significantly to the cost of producing natural gas. Currently, most high nitrogen reserves are not produced because the market price of gas at the pipeline is not sufficient to pay for nitrogen rejection. Cryogenic processes have been thoroughly optimized and there is little prospect that the cost of cryogenic separations can be significantly reduced.

Cryogenic processes are inherently expensive because of the very low temperatures encountered. Feeds must be purified to very low concentrations of water, carbon dioxide and mercury because minute amounts of these impurities have deleterious effects at cryogenic temperatures. Exotic, costly materials are required to withstand cryogenic temperatures. The intensive heat exchange integration required to reduce energy consumption makes cryogenic units difficult

to control and operate and impossible to adapt to changing feeds and product slates, which are often required in energy markets.

For the foregoing reasons, there is need for a cost effective, flexible and safe process for rejection of nitrogen from natural gas to allow production of our vast but dormant high nitrogen natural gas resources.

Hydrogen Recovery from Refining and Petrochemical Off-Gases

Demand for hydrogen to desulfurize feedstocks and liquefy heavy fractions in refining and petrochemical operations is increasing because sulfur in crude oils now being refined is increasing and more heavy crudes with higher carbon-to-hydrogen ratio are being produced. Most of the required chemical hydrogen is made on purpose by steam reforming or partial oxidation of methane and other hydrocarbons. It is estimated, however, that refining and petrochemical plants in North America are burning offgases containing 1,500 to 2,000 million cubic feet per day of recoverable hydrogen. Several major off-gas sources of hydrogen are listed in Table 2. Apparently, even though the chemical value of hydrogen is two to three times its fuel value, it is cheaper to make on-purpose hydrogen than to recover hydrogen in these off-gas streams using current technology (cryogenic, membrane, and pressure swing adsorption processes).

For the foregoing reasons, there is need for a cost effective, flexible and safe process for recovering hydrogen from refinery and petrochemical off-gases that will make recovered hydrogen cost competitive with on-purpose hydrogen.

Mehra ProcessSM

Mehra processes are absorption processes that utilize a physical absorption solvent to separate and recover hydrogen, methane, ethylene and other valuable hydrocarbons from mixed hydrocarbon streams. Mehra technology has been applied to recovery of ethylene and hydrogen from refinery and petrochemical off-gas streams and to reject nitrogen from natural gas. Depending on the application,

Mehra specifies absorption solvents that are selected from preferred groups to optimally synergize with the process. Among the preferred Mehra solvents are C4 to C10 hydrocarbons including paraffins olefins and aromatics. Mehra technology is described in U.S Patents Nos. 4,832,718 and 4,740,222, which are incorporated
5 herein by reference.

In general, Mehra processes operate at a higher temperature than cryogenic processes which provides advantages over cryogenic processes: 1) Exotic cryogenic construction materials required to withstand cryogenic temperatures are
10 not required in Mehra processes; 2) Feed purification specifications are more relaxed; 3) Cryogenic processes are intensively heat integrated to reduce energy consumption whereas Mehra processes are not. Accordingly Mehra processes are more flexible and adaptable to changes. Process conditions can be changed quickly "on-line" with no adverse impact on operability and without equipment
15 modifications to alter product stream compositions or maintain product composition should feed composition change.

The present invention is directed to an absorption process utilizing Mehra technology for separating a feed stream consisting of components with a spectrum
20 of volatilities components ranging from volatile components such as hydrogen and nitrogen to intermediate volatility components such as methane, ethylene and ethane to heavier hydrocarbons in the C3 to C10 carbon number range. In one embodiment, the process of this invention can separate the feed into two streams: a light product stream consisting predominantly of volatile (light) component and a
25 heavy product stream consisting predominantly of less volatile (heavy) components. Alternatively, the process of this invention can separate the feed stream into three product streams: a light product stream consisting predominantly of light components, an intermediate product stream consisting predominantly of feed components with intermediate volatilities, and a heavy product stream
30 consisting predominantly of heavy or low volatility components. The volatility cut points between the product groups can be varied up or down in volatility among

components with adjacent volatility by changing operating conditions, e.g., the least volatile component of the most volatile group can become the most volatile component of the intermediate group.

5 The process of the present invention comprises the steps of:

 (a) countercurrently contacting the feed gas stream with an absorption solvent stream in the absorption section of an absorber stripper to form a light product gas stream comprised of light components and a rich solvent bottoms stream comprised of solvent and absorbed intermediate and heavy components.

 (b) flashing the rich solvent absorber bottoms stream from the to reduced pressure in one or more flash drums to produce a flash gas stream and a liquid stream and conveying the liquid stream to the absorber stripper as the absorption solvent; and

 (c) wherein the absorption solvent is consists essentially of heavy components that are derived from the feed; no external solvent is used.

20 Another important aspect of the present invention are the techniques used to control the inventory and composition of the heavy hydrocarbon absorption solvent.

 The preceding and other features, aspects and advantages of the present invention will become apparent from the following description, appended claims, and accompanying drawings.

 Figure 1 is a simplified flow diagram of an embodiment of the present invention which separates a feed stream comprised of light, intermediate and heavy components into a light product stream comprised of light components and a heavy product stream comprised of intermediate and heavy components.

Figure 2 is a simplified process flow diagram of an embodiment of the present invention which separates a feed stream comprised of light, intermediate and heavy components into a light product stream comprised of light components, an intermediate stream comprised of intermediate components and a heavy product stream comprised of heavy components.

Figure 3 is a simplified process flow diagram of an embodiment of this invention which includes additions to the process of Figure 2 to reduce the concentration of intermediate and heavy components in the light product stream.

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In the interest of clarity, pumps, heat exchangers, control valves, control systems, and auxiliary equipment items that are necessary for a practical and safe operating unit have intentionally been left out if not required to elucidate the inventive concept. These deletions are generally understood by those skilled in the art and do not limit the scope of the invention.

15

Prior to entering the process, the feed stream is dried to preclude formation of solid hydrates in the process. The water dewpoint specification is typically in the range of about -35°C (-30°F) to about -100°C (-150°F) which can be readily achieved with conventional drying technology such as a molecular sieve or glycol injection drying system; the lower the minimum temperature in the process, the lower the dewpoint specification.

20

The flowsheet of Figure 1 depicts an embodiment of the present invention in which the feed is separated into a light product stream consisting predominantly of light components and a heavy product consisting predominantly of heavy components. The feed gas 10 is introduced into the bottom of an absorber column 19 and flows up the column. Lean solvent is introduced near the top of the column and flows down the column countercurrently contacting the gas stream. The absorber is packed with Raschig rings or other packing 21 with large surface area and low pressure drop that promotes liquid/gas mass transfer. Alternatively,

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distillation trays such as sieve or valve trays can be used to affect the required mass transfer.

5 The solvent absorbs the intermediate and heavy components from the feed stream. The overhead gas stream leaving the top of the absorber 24 is the light product stream which consists predominantly of light components. The bottoms stream 30 is rich solvent which contains the heavy components absorbed out of the feed gas stream.

10 The solvent is regenerated by flashing to a lower pressure in one or more flash drums in series. (One flash drum 37 is shown in Figure 1.) The flash gas coming overheads from the flash drums 38 contains intermediate and heavy components from the feed. The flash gas stream 38 is cooled in the product cooler 47 to condense the heavy components. The condensate is separated from
15 the flash drum overhead stream in the product separator 49 and conveyed to the flash drum through line 50. The gas stream from the product separator 48 is called the heavy product stream and it contains intermediate and heavy components that came in with the feed.

20 The bottoms stream from the flash drum 39 is the lean absorption solvent stream and it is pumped into the top of the absorber. The solvent consists predominantly of heavy components derived from the feed. The temperature of the lean solvent stream 43 is reduced in a solvent cooler 18 before it is conveyed into the top of the absorber 19.

25 It is important to minimize heavy components lost in the absorber overhead gas stream. Measures which improve absorption of heavy components include, in the design phase, providing more mass transfer stages--more mass transfer surface area and higher efficiency--in the absorption section. Operating measures include
30 increasing lean solvent circulation rate relative to gas feed rate, and reducing the temperature of the lean solvent stream coming into the tower. Obviously, these

measures raise costs, so there is an economic trade-off between minimizing overhead losses and optimizing cost.

5 An inventory of solvent is maintained in the flash drum 37. If less heavy components come in with the feed than leave in the light and heavy products streams, the solvent inventory will decrease with time. If more heavy components come into the absorber in the feed than leave in the product streams, then the solvent inventory will increase with time. The solvent inventory is stabilized by
10 controlling the temperatures at the outlet of the solvent cooler 18 and the product cooler 47 to make the heavy components coming into the absorber in the feed equal to the heavy components leaving in the product streams. The heavy product stream 48 is saturated with heavy components so increasing its temperature increases heavy components leaving the process in stream 48 and reducing the temperature of stream 48 reduces heavy components leaving in the stream 48.
15 The light product stream 24, is also saturated with heavy components. Reducing the temperature of the solvent going into the absorber 20 reduces the temperature at the top of the absorber which, in turn, reduces the amount of heavy components in the light product stream.

20 Figure 2 is a simplified flow plan of an embodiment of the present invention in which three product streams are produced: a light product stream consisting predominantly of light components; an intermediate product stream consisting primarily of intermediate components; and a heavy product stream consisting predominantly of heavy components.

25

The feed stream 10 is cooled in feed cooler 11 to condense a heavy product stream 14 which consists of predominantly heavy components. The heavy product condensate stream 14 is separated from the feed stream 10 in a feed separator 13. The residual gas phase 17 is conveyed to the absorber stripper 19. Typically, to
30 conserve refrigeration energy, the feed coming into the process is cooled by heat exchange against the cold light product stream.

The overhead stream from the feed separator 17 is conveyed into an absorber stripper 19. Lean solvent which has been cooled in the solvent cooler 18 is conveyed into the top of the absorber stripper through line 20.

5 The absorber stripper 19 contains packing 21 which promotes liquid/gas mass transfer exchange, such as Raschig rings or saddle packing. Alternatively, distillation trays such as bubble caps, sieve trays, or high efficiency trays with multi downcomers can be used to affect liquid gas mass transfer exchange in the tower.

10

 The section of the absorber stripper above the feed point is the absorption section 22. In the absorption section 22, the rising feed gas stream countercurrently contacts the falling solvent stream. Intermediate and heavy components are absorbed out of the feed gas stream into the solvent stream. Most
15 of the light components in the feed are not absorbed by the solvent in the absorption section 22 so that the absorber stripper overhead stream 24 contains most of the light components which were in the feed stream 10.

 The absorber stripper 19 operates at feed line pressure, typically 150 to
20 3000 psi. In some applications, the light product exits the process at absorber stripper pressure through stream 24A. In other applications, the light product stream is vented to the atmosphere via line 24. For these applications the stream 24 pressure can be reduced across a control valve, or, as shown in Figure 2, through an expander 25 to recover shaft power from the expanding gas. The
25 temperature of the absorber stripper overhead gas stream 26 drops sharply as its pressure is reduced. Temperature drops more through an expander than a control valve because, through an expander, the expansion is isentropic with work extracted from the expanding gas, whereas, across a valve, the expansion is adiabatic. In either case, the temperature reduction causes heavy and intermediate
30 components in the light product to condense. The condensate is separated in the light product separator 27 and is conveyed into a flash drum via line 28.

The absorption solvent stream 42 is produced as the bottoms stream from the fourth flash drum 43 and it consists predominantly of heavy components derived from the feed, no external solvent is used in the process.

5 Some light components are absorbed by the solvent in the absorption section 22 of the absorber stripper 19 and are carried into the stripping section 23. These light components leave the process in the intermediate product stream if they are not stripped out of the stripper absorber bottoms stream 30.

10 In some applications, the concentration of light components in the intermediate products stream must be very low to meet intermediate product specifications. Accordingly, Figure 2 includes a stripping section 23 in the absorber 19 below the feed point. In the stripping section 23 light components are stripped from the solvent while retaining intermediate and heavy components in the
15 solvent. This is accomplished by countercurrently contacting the rich solvent as it cascades down the stripping section with a stripping gas stream that flows up the column. The stripping gas stream is formed by flashing the absorber stripper bottoms stream 30 from tower pressure to a lower pressure across valve 31 into first absorber stripper bottoms flash drum 34. The stripping gas stream is taken
20 overhead from the first flash drum 33, compressed to tower pressure using a suitable compressor 35, such as a reciprocating or centrifugal compressor, and conveyed into the absorber stripper tower near its bottom as stream 36.

25 Light component stripping effectiveness can be increased by heating the stripping gas stream 36 in a heater (not shown in Figure 2). Increasing flow of stripping gas to the absorber stripper, by increasing the speed of the compressor 33, also increases stripping efficiency.

30 The bottoms stream 34 from the first flash drum 32 is flashed to reduced pressure through three flash drums 37, 40 and 43, connected in series. The pressure in the flash drums decreases moving downstream. Pressure can be

reduced across pressure letdown valves. Alternatively, expander turbines can be employed in lieu of pressure letdown valves to recover energy from the flashing streams. The number of flash stages can be varied; the optimal number depends on feed and products specifications and the cost of adding and operating additional flash drums. The flash gas streams from the flash drums are compressed to intermediate product pipeline or storage pressure in a compressor 46 and combined to form the intermediate product stream.

The intermediate product compressor 46 can be either a reciprocating or centrifugal compressor. However, a multi-stage compressor is preferred. Overhead gas from the second (upstream) flash drum 37, which is at the highest pressure, is fed into the suction of the high pressure (downstream) compressor stage 38; overhead gas from the fourth (downstream) stage 43, which is at the lowest pressure, is fed into the suction of the low pressure (upstream) compressor stage 44; overhead from intermediate flash stages 40 are fed into intermediate compressor stages 41. Use of multiple flash drums at intermediate pressures reduces the size and power consumption of the intermediate product gas compressor 46 relative to using only one flash drum and feeding all the overhead flash gas into the suction of the compressor, since not all the flash gas has to be compressed up from the lowest downstream flash drum pressure to discharge pressure.

The intermediate product compressor discharge stream 51 is cooled in the intermediate product cooler 47 to condense heavy components. The heavy component condensate is separated in the intermediate product separator 49 and conveyed to a flash drum via line 50. The overhead gas stream 48 from the intermediate product separator 49 is the intermediate product stream.

The bottoms stream 45 from the fourth (downstream) flash drum 43 is the lean absorption solvent stream and it is pumped from the fourth flash drum 45 through the solvent cooler 18 to reduce its temperature and then into the top of the

absorber stripper tower stream via line 20. The solvent consists predominantly of heavy components derived from the feed stream 10.

The intermediate product compressor discharge stream 51 contains heavy components. To control the amount of heavy components in the intermediate product, stream 51 is cooled in an intermediate product cooler 47 to condense heavy components. The condensed heavy component stream is separated from the intermediate product stream in an intermediate product separator 49 and conveyed via line 50 to a flash drum.

Solvent inventory in the flash drums is monitored. If more heavy components leave in the product streams than come into the process, over time, solvent inventory in the flash drums will be depleted. If liquid inventory in the flash drums falls below the setpoint, solvent must be replenished and/or heavy component losses reduced. One way to make up solvent inventory is to divert some of the heavy product stream 14 to the flash drums via line 16. Alternatively or additionally, the exit temperature from the feed cooler 11 can be increased which will admit more heavy components into the absorber stripper 23 via line 17.

Losses of heavy components in the intermediate product stream need to be limited to avoid depleting solvent inventory. This is accomplished by maintaining the temperature at the outlet of the intermediate product cooler 47 sufficiently low to condense heavy components out of the intermediate product stream.

Also, heavy and intermediate components leave the process in the light product stream 24 or 24A. These component losses are limited by reducing the temperature of the lean solvent at the outlet of the solvent cooler 18, which reduces the temperature at the top of the absorber stripper 19, which, in turn, reduces the vapor pressure of intermediate and heavy components at the top of the absorber stripper in the light product gas stream 24 or 24A. The bottoms stream 28 from the light product separator 27, which contains heavy and intermediate

components, and the bottoms stream 50 from the intermediate product separator 49, which contain heavy components are to a flash drum 43.

Commonly, a refrigerant such as boiling freon, propylene or propane is specified as the cooling medium in the intermediate product cooler 47, the solvent cooler 18, and the feed cooler 11 to attain exchanger outlet temperatures sufficiently cold to reduce intermediate and heavy component losses. Product cooler effluents are typically maintained at the coldest practicable temperature.

If solvent inventory in the flash drums increases with time and exceeds the maximum setpoint the actions discussed above to avoid depleting solvent are reversed.

In the ideal situation, it is possible to adjust outlet temperatures from the intermediate product cooler 47, the feed cooler 11, and the solvent cooler 18, to make the heavy components leaving the process equal the rate the heavy components coming into the absorber stripper, via stream 17, which will stabilize solvent inventory.

Note that in the process of Figure 2 the solvent stream 42 produced in the flash drums is comprised essentially of heavy hydrocarbons that are derived from the feed stream.

The process of Figure 3 is another embodiment of the present invention. The process of Figure 3 is similar to the process of Figure 2 except that it includes one more tower, the solvent fractionator 60, and utilizes two lean solvent streams, a main lean solvent stream 63 and a slip lean solvent stream 62. The added equipment and complexity reduce the concentration of intermediate and heavy components in the light product stream 24 or 24A relative to the process in Figure 2 and is therefore useful when a light product of exceptional purity is required.

The solvent fractionator 60 produces a highly purified lean solvent slip stream as fractionator bottoms 61 by fractionating part of the bottoms stream from the final downstream flash drum 64. The lean solvent slip stream 61 flow rate is typically about 15% to 70% of the main solvent stream flow rate 63. The lean solvent slip stream 61 contains substantially less intermediate components than main lean solvent stream 63. The lean solvent slip stream is cooled in solvent cooler 64 and conveyed into the top of the absorber stripper via line 62 where it countercurrently contacts the absorber stripper overhead just before the overhead stream exits the tower. The main solvent stream 63 is conveyed into the absorber stripper 19 via line 63 at a point below the slipstream inlet point on the tower midsection.

The overhead stream from the solvent fractionator 65 is predominantly comprised of intermediate components and is added to the intermediate product stream 48.

Contact with the highly purified slip stream solvent in the top of the absorber stripper produces the desired exceptionally pure light product gas stream 24 or 24A.

Note that in the process of Figure 3, both the main solvent and the slip stream solvent are comprised predominantly of heavy hydrocarbons that are derived from the feed.

When initially starting up a process of this invention, there is of course no solvent inventory in the flash drums. The required solvent inventory can be accumulated by operating the unit with no lean solvent flow to the absorber stripper until liquid level in the flash drums reaches the set points. The composition of the solvent in the flash drums will, over a period of a few hours, equilibrate to its steady state composition. Alternatively, heavy product condensed out of the feed, stream 16, can be totally diverted to the flash drums until the

solvent levels reach the setpoint. This batch of solvent can be a mixture of components blended to have a composition similar to the composition of the heavy components, or heavy component condensate obtained from the feed for the plant, or it can be one of the heavy components. Whatever the initial solvent
5 composition, the composition will, over a period of several hours of operation, equilibrate to a steady state composition consisting primarily of heavy components in the feed, as determined by the operating conditions in the unit, and no external solvent will be required to replenish solvent losses.

10 **Application to Natural Gas Nitrogen Rejection**

An important commercial application of the present invention is rejection of nitrogen and helium from natural gas. In natural gas rejection applications, the natural gas (methane) product stream is the intermediate product stream. The nitrogen and other non-condensable gases are rejected in the light product stream.
15 The natural gas is compressed and delivered into a natural gas transmission pipeline.

Typically, the pipeline operators require that the concentration of nitrogen and other non-combustible gases in the natural gas be reduced to between 1 mole
20 % and 10 mole %, more preferably between 2 mole % and 5 mole %, and most preferably between 2 mole % and 3 mole %. Usually, energy market conditions dictate that a heavy product stream consisting predominantly of ethane and heavier hydrocarbons be produced. The heavy product stream is produced by cooling the feed stream to condense heavy components. The heavy product stream is
25 separated from the feed stream and the feed stream is conveyed to the absorber. The heavy stream is usually sold as an LPG fuel, or for chemical feed stock material.

An advantage of the process of the present invention is that absorption
30 tower operates at natural gas feed line pressure which is typically in the range of 150 to 3000 psia. Accordingly, the rejected nitrogen stream is produced at

pressures high enough that the stream can be re-injected into well formations to enhance recovery of oil and gas.

Helium contained in the natural gas feed appears in the light product
5 (rejected nitrogen) stream. If helium is present in the light product stream in sufficient quantity to cost-effectively recover it, the light product stream is conveyed to a helium recovery unit before discarding it. Helium recovery in conjunction with Mehra processes is disclosed in U.S. Patent No. 5,224,350, which is incorporated herein by reference.

10

Application to Hydrogen Recovery

Another important commercial application of the present invention is recovery of hydrogen from refinery and petrochemical plant off-gas streams. In hydrogen recovery applications, hydrogen is recovered in the light product stream.
15 If, as is commonly the case, the hydrogen feed stream contains methane, then the intermediate product stream produced is a methane-rich stream that is suitable for fuel gas. Usually, energy product market conditions dictate that a heavy product stream be produced from the feed gas. The heavy product stream is produced by cooling the feed stream to condense heavy components which include ethylene,
20 ethane and heavier hydrocarbons. The heavy product stream is separated from the feed stream and the feed stream is conveyed to the absorber. The heavy product stream is sold as an LPG fuel or used as a chemical raw material. If it is not cost effective to produce a heavy product stream, then the heavy components in the feed leave the process in the intermediate product stream.

25

Non-condensable impurities including CO and nitrogen are commonly present in hydrogen off-gases. Non-condensables appear in the light product stream along with the hydrogen. Many hydrogen applications require that non-condensable impurities in the hydrogen be reduced to very low concentrations to
30 yield hydrogen streams as pure as 99.995 mole %. For these applications, the

light product stream coming overhead from the absorber is conveyed to a suitable hydrogen purification unit such as a pressure swing adsorption (PSA) system.

Hydrogen bearing off-gases from ammonia plants contain nitrogen. The
5 nitrogen leaves the process in hydrogen light product stream in the same ratio as nitrogen to hydrogen in the feed. Typically, the ratio of nitrogen to hydrogen in the feed is too high for most applications so that the nitrogen to hydrogen ratio in the light product stream must be increased. This can be accomplished in any of the commercial hydrogen purification systems such as a PSA unit.

10

Example One: Nitrogen Rejection from Natural Gas

Table 1 is a material balance which shows how the process of Figure 2 is applied to reject nitrogen from natural gas. (The stream numbers in Table 3 refer to Figure 2).

15

The feed rate is 40 MMSCFD of natural gas containing 30 mol % nitrogen and 11 mol % C₂+ hydrocarbons. The feed also contains small amounts of helium and carbon dioxide. The unit reduces nitrogen content of the natural gas product to 2 mol % while limiting methane losses in the rejected nitrogen to only
20 2% of incoming methane.

The light products stream 29 contains the rejected nitrogen and also the helium present in the feed. The intermediate product stream 48 is the natural gas product stream, suitable for transmission in the natural gas pipeline grid. The
25 heavy product stream 15 is predominantly C₂ to C₆ hydrocarbons which can be sold as LPG fuel or for chemical raw material, or be blended into the natural gas product stream 48.

The feed stream 10 is dehydrated to a water dewpoint of -100°C (-150°F) in
30 a mole sieve dryer unit before coming into the rejection plant. The dry feed 10 enters the nitrogen rejection unit at about 265°C (80°F) and about 700 psig.

The feed stream is cooled to -26°C (-14°F) in feed cooler 11 by heat exchange with product streams which exit the process at very cold temperatures. (Feed/product heat exchange cost effectively reduces refrigeration energy consumption). C_2+ hydrocarbons condense out of the feed as it is cooled in feed
5 cooler 11. The condensate is the heavy product 15. It is separated from the feed stream in feed cooler separator 13 as the feed cooler separator bottoms stream 14.

The cooled feed stream 17 is conveyed into the absorber stripper 19 at a point near the bottom of the absorption section 22 and flows upward
10 countercurrently contacting the solvent stream 20. The absorber stripper operates at feed line pressure, 700 psia.

The solvent stream 20 enters into the absorber stripper at the top of the tower. It is the bottoms stream from the fourth flash drum 45 which has been
15 cooled in solvent cooler 18, and it consists predominantly of C_2+ components, which are the heavy components of the feed.

The rejected nitrogen and helium is contained in absorber stripper overhead stream 24.
20

In this application, the rejected nitrogen light product stream 24 is vented to the atmosphere. Rather than reducing pressure of the absorber stripper overhead stream 24 across a control valve, it is cost effective to reduce the pressure, from 700 psia to 43 psia, in the light product expander 25 and extract
25 shaft power. The temperature of the gas stream falls to about -137°C (-215°F) as it isentropically depressurizes in the expander so that the intermediate component, methane, and heavy components, C_2+ hydrocarbons partially condense. The condensate is separated from the nitrogen stream in the light product separator as the bottoms stream 28 and recycled back into the process downstream. The cold
30 rejected nitrogen stream 29 coming overhead from the light product separator, which is at about -137°C (-215°F), is cross heat exchanged with incoming feed in

feed pre-coolers (not shown on Figure 1), and is vented to the atmosphere at about 24°C (74°F).

In other applications, rather than venting the rejected nitrogen stream to the atmosphere, the nitrogen is conveyed back down into a subterranean oil or gas formation to enhance production. Clearly, at locations where nitrogen injection production enhancement is advantageous one advantage of the Mehra nitrogen rejection process over cryogenic processes is that the rejected nitrogen stream is produced at high pressure suitable for injection without compression.

The C2+ solvent stream from the bottom of the absorption section 22 of the absorber stripper, enriched with absorbed methane, flows down through the stripping section 23 of the absorber stripper where it is countercurrently contacted with stripping gas 36. The function of the stripping section is to strip nitrogen and other light components out of the solvent stream. Since nitrogen in the absorber stripper bottoms eventually exits the process in the natural gas product stream, the molar ratio of nitrogen to methane in the absorber stripper bottoms must be reduced below the maximum concentration of nitrogen and other non-combustibles allowed by the gas pipeline system that will transport the gas. In this example, the molar ratio of nitrogen to methane in the bottoms stream 34 from the first flash drum is reduced to about 2%, which will make the mole % of nitrogen in the natural gas intermediate product 48 less than 2%.

The stripping gas stream is produced by flashing the absorber stripper bottoms stream 20 to a pressure of 360 psia in the first flash drum 32. The overhead stream from the first flash drum 32, which is the stripping gas stream, is compressed up to stripper absorber pressure of 700 psia using stripping gas compressor 33, then passed through a heat exchanger (not shown on Figure 1), which adjusts the stream temperature to about 35°C (94°F), and into the bottom of the absorber stripper. The stripping gas flow rate is 2420 lb/mol per hour.

In this example, three stages of flashing are used to regenerate the rich solvent and recover the natural gas: flash drums 37, 40, and 43, arranged in series. Three flash stages is the economic optimum for this application. For other applications, fewer or more stages may be optimal. The flash drums operate at successively lower pressures, 245 psia in the flash drum 37, 70 psia in the flash drum 40, and 20 psia in flash drum 43, respectively. The flash drum overhead streams 38, 41, and 44 contain the natural gas. They are compressed to natural gas pipeline pressure, 815 psia, in a four-stage reciprocating compressor 46. The upstream low pressure compressor stage takes suction from downstream flash drum overhead stream 44, the downstream high pressure compressor stage takes suction from the upstream flash drum overhead stream 38, and the intermediate compressor stage takes suction from the middle flash drum overhead stream 41. The fourth compressor stage takes suction from the discharge of the third stage and boosts the discharge pressure to pipeline pressure.

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The compressed natural gas stream is cooled in the intermediate product cooler 47 to about -23°C (-9°F) using propane refrigerant coolant. Heavy, C_2+ , components condense in the intermediate product cooler 47. The condensate is separated from the natural gas in the intermediate product separator drum 49 and conveyed from the separator to the downstream flash drum 40 using line 50.

20

The bottoms stream 45 from the fourth stage flash, which consists predominantly of heavy, C_2+ , components derived from the feed, is the lean solvent stream. The stream is pumped through the solvent cooler 18, where it is cooled to -30°C (-25°F) (using propane refrigerant as the cooling medium), and then enters into the absorber stripper 19 at a point near its top via line 20.

25

The bottoms streams from the light product separator 27 and the intermediate product separator, which contain heavy (C_2+) components, are conveyed to the fourth flash drum to reduce depletion of solvent inventory in the flash drums. The solvent inventory in the four flash drums is continually

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monitored. In this example, solvent inventory in the flash drums is stabilized by controlling the temperatures at the exit of the feed cooler 11, the solvent cooler 18, and the intermediate product cooler 47, to make heavy components going into the absorber stripper via stream 17 equal the heavy components leaving the process in the nitrogen rich light product stream 29 and the methane rich intermediate product stream 48. Heavy product make-up to replenish the inventory is not required.

If the C2+ components lost in the rejected nitrogen and natural gas product streams were to exceed the C2+ components coming into the process in the feed to the absorber stripper 19, solvent would be continuously depleted. To replenish solvent inventory, heavy product from the feed separator 13 could be diverted to one of the flash drums via line 16 to replenish the solvent inventory. Alternatively, to equilibrate heavy component losses in the light and intermediate product streams to heavy components in the feed to the absorber stripper tower, one could raise the temperature at the exit of the feed cooler 11 which would bring more heavy components into the absorber stripper 19 or reduce the temperature at the exit of the intermediate product cooler 47, which would reduce the loss of heavy components in the natural gas intermediate product stream, or reduce the temperature at the exit of the solvent cooler 18, which would reduce the loss of heavy components in the nitrogen rich light product stream, or any combination of these temperature adjustments to replenish and stabilize solvent inventory.

An energy saving option which can be cost effective is to replace the flash valves in lines 30, 34, 39 and 42 with hydraulic turbines which extract shaft work out of the expanding flashing streams which is converted to power. In addition, lower flash temperatures are achieved using turbines instead of valves which reduces the requirement for external refrigerants in the refrigerated heat exchangers used in the process.

Example Two: Hydrogen Recovery From Refinery Off-Gas

An overall material balance indicating how the process of Figure 2 may be applied to recover hydrogen from a typical petroleum refinery offgas stream is presented in Table 4. (The stream numbers in Table 4 refer to Figure 2). The feed stream 10 contains 20 mole % hydrogen. (It is currently being burned as fuel gas.) The process of this invention recovers a light product stream containing 75 mole % hydrogen, an intermediate product stream that is predominantly methane and is a suitable fuel gas, and a heavy product stream that contains C2 to C5 components, the heavy components of the feed, which can be sold for fuel or as a chemical raw material or blended into fuel gas intermediate product stream. In this example, the heavy components are blended into the intermediate product fuel gas stream 48.

The dominant heavy component in the feed stream 10 is n-pentane. Accordingly, the solvent stream 20 is also rich in n-pentane, 37 mole % n-pentane. No external solvent is used; the solvent consists predominantly of heavy components derived from the feed.

The solvent inventory, accumulated in the flash drums, is maintained by controlling the temperatures at the outlet of the feed cooler 11, solvent cooler 18, and the intermediate product cooler 47 so that the heavy components in the feed going into the stripper absorber in stream 17 equals the heavy components in the light and intermediate product streams.

In this example, 99.8% of the hydrogen in the feed is recovered in the light product hydrogen rich stream 24A. A light product expander is not used because the hydrogen rich light product stream is delivered to its application at full absorber stripper pressure, so there is no pressure letdown.

The processes of the present invention depicted in Figures 3 and 4 recover ethylene, hydrogen, methane and heavy hydrocarbons from petrochemical and

refinery off-gas streams via physical absorption using solvents comprised of heavy hydrocarbons that are derived from the feed and produced within the process; no external solvent is used. molecular weight range by controlling the pressure in the flash drums.

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In conventional absorption processes, the solvent is an external solvent that is purchased and stored in a tank near the absorber stripper. Periodically or continually, solvent is pumped from the storage tank to the absorber stripper to makeup losses of solvent. The process of this invention does not use an external
10 solvent, but instead produces the solvent from heavy components in the feed stream. Accordingly, there is no need to purchase, store and handle external solvent. This results in significant reduction in investment and operating costs relative to processes that use external solvents.

15 The process of the present invention sufficiently is cost competitive, flexible, and safe to induce operators to recover ethylene and hydrogen from refining and petrochemical off-gas rather than to burn them in flares or consume them as relatively low value fuel gas. Clearly, the prospects for penetration and acceptance of absorption processes for recovering ethylene and associated
20 hydrogen from refinery and petrochemical off-gas streams are enhanced by the improvements encompassed in the process of the present invention.

The embodiments of the process of this invention discussed above are representative embodiments of the invention. Those skilled in the art will
25 recognize that many other process designs can be devised that embody the present invention.

The process of the present invention is a practical, operable, flexible, adaptable and cost effective technology for rejecting nitrogen from natural gas.
30 The process can produce vast reserves of high nitrogen natural gas that currently can not be produced with available cryogenic technology.

The process of the present invention can also be applied to recover hydrogen from refinery and petrochemical off-gas streams which is currently wasted because with currently available cryogenic processes it costs more to recover the vented hydrogen than to produce and purchase makeup hydrogen.

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Clearly, the prospects for penetration and acceptance of absorption processes for producing high nitrogen natural gas and recovering hydrogen from refinery and petrochemical off-gas streams are enhanced by the improvements encompassed in the process of the present invention. In conventional absorption processes, the absorption solvent is an external solvent that is purchased and stored in a tank near the absorber stripper. Periodically or continually, solvent is pumped from the storage tank to the absorber stripper to makeup losses of solvent. The process of this invention does not use an external solvent, but instead produces the solvent from heavy components in the feed stream. Accordingly, there is no need to purchase, store and handle external solvent. This results in significant reduction in investment and operating costs relative to processes that use external solvents. The costs and equipment associated with buying, receiving, storing and charging makeup solvent are eliminated. These are particularly significant advantages for natural gas treatment units which are usually located at remote sites. Accordingly, the technical advantages of the present invention will increase the rate of penetration of the new absorption technology into natural gas production and hydrogen recovery markets.

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TABLE 1
Typical Natural Gas Compositions

Component	Symbol	Mol Percent		
		Associated Gas	Wet Gas	Dry Gas
Methane	C ₁	27.52	59.52	97.17
Ethane	C ₂	16.34	5.36	1.89
Propane	C ₃	29.18	4.71	0.29
i-Butane	i-C ₄	5.37	2.03	0.13
n-Butane	n-C ₄	17.18	2.39	0.12
i-Pentane	i-C ₅	2.18	1.80	0.07
n-Pentane	n-C ₅	1.72	1.61	0.05
Hexane	C ₆	0.47	2.60	0.04
Heptanes Plus	C ₇₊	<u>0.04</u>	<u>19.98</u>	<u>0.24</u>
		100.00	100.00	100.00

TABLE 2
Sources of Hydrogen Off-Gas Streams

Industry	Source	Approximate Hydrogen Concentration (Mol%)
Refining	HT Purge	25-35
	FCC Gas	10-15
	Cascade Reject	50-60
Methanol	Purge Gas	70-80
Ethylene	By-product H ₂	60-90
	Cracked Gas	10-35
Coke Oven	Product Gas	0-5
LPG Dehydrogenation	Product Gas	58
Toluene HDA	H ₂ Purge	57
Cyclohexane	H ₂ Purge	42
Carbon Black	Product Gas	7
Formaldehyde	By-product H ₂	18
Ammonia	Purge Gas	60

TABLE 3A
Example 1 (Figure 1)
Natural Gas Nitrogen Rejection 40 MMSCFD Feed

Stream Name	Feed	Feed Separator Bottoms/ Heavy Product	Feed Separator Overhead	Absorber Stripper Overhead	Light Product	First Flash Drum Overhead	First Flash Drum Bottoms	Second Flash Drum Overhead	Third Flash Drum Overhead	Fourth Flash Drum Overhead	Fourth Flash Drum Bottoms	Intermediate Product Separator Bottoms	Intermediate Product
Stream Number (Figure 1)	10	14	17	24	29	33	34	38	41	44	45	50	48
Temperature, °F	80	-14	-14	-24	74	-8	-8	-11	-19	-24	-20	-9	120
Pressure, PSIA	715	712	712	700	40	360	360	245	70	20	20	444	815
Mass Flow, LB/HR	97,770	6,638	91,132	39,917	36,104	43,770	1,014,347	15,979	29,873	14,045	960,129	4865	55,031
Gas Flow, MMSCFD	40	--	38.66	12.13	11.96	22.04	--	8.16	14.23	5.3	--	--	26.7
Liquid Flow, GPM	--	23.56	--	--	--	--	3,165.04	--	--	--	2,837.71	17.47	--
Composition, LB MOL/HR													
Helium	3.95	--	3.95	3.94	3.94	0.04	--	0.01	--	--	--	--	0.01
Nitrogen	1,317.66	4.87	1,312.79	1,253.84	1,253.58	205.69	58.91	39.51	18.37	1.48	0.08	0.15	59.21

TABLE 3B
Example 1 (Figure 1)
Natural Gas Nitrogen Rejection 40 MMSCFD Feed

Stream Name	Feed	Feed Separator Bottoms/ Heavy Product	Feed Separator Overhead	Absorber Stripper Overhead	Light Product	First Flash Drum Overhead	First Flash Drum Bottoms	Second Flash Drum Overhead	Third Flash Drum Overhead	Fourth Flash Drum Overhead	Fourth Flash Drum Bottoms	Intermediate Product Separator Bottoms	Intermediate Product
CO ₂	4.39	0.15	4.24	0.14	0.12	1.43	5.63	0.67	2.15	1.41	1.52	0.11	4.12
Methane	2,602.36	33.70	2,568.68	51.63	51.34	2,112.85	2,624.10	808.81	1,347.82	384.45	107.51	23.74	2,517.34
Ethane	256.51	20.04	236.47	10.63	4.63	56.33	564.88	27.42	113.25	106.46	338.53	15.29	231.84
Propane	106.73	26.47	80.26	5.78	0.09	21.71	1,004.52	10.17	43.35	48.77	928.97	22.12	80.17
i Butane	29.43	13.50	15.93	1.84	--	6.76	917.70	300	11.71	13.32	903.10	12.10	15.93
n Butane	36.89	20.52	16.37	2.42	--	9.03	1,893.99	3.00	14.67	16.78	1,879.17	19.00	16.37
i Pentane	12.30	9.34	2.96	0.75	--	2.87	1,729.73	1.18	4.69	4.45	1,727.96	7.36	2.96
n Pentane	10.99	8.97	2.02	0.46	--	1.80	1,589.74	0.72	3.42	2.61	1,589.06	4.73	2.02
n Hexane	10.98	10.30	0.68	0.38	--	1.67	5,873.19	0.63	2.33	1.90	5,873.29	4.18	0.68
								--	--	--	--		
TOTAL, LB MOL/Hr	4,392.20	147.85	4,244.35	1,331.83	1,313.71	2,420.17	16,262.39	896.05	1,561.76	581.62	13,349.19	108.78	2,930.65

-28-

TABLE 4

Example 2

Hydrogen Recovery (Figure 1)
Overall Material Balance

Stream	Feed	Light Product	Intermediate Product	Solvent
Stream Number	10	24A	48	20
Temperature, °F	100	95	74	-25
Pressure, PSIA	325	307	75	310
Composition, <u>LB-MOL/HR</u>				<u>MOL%</u>
Hydrogen	1,103.02	1,100.41	2.60	
Nitrogen	108.89	90.56	18.33	
H ₂ S	15.25	0.74	14.51	
Methane	3,255.13	213.54	3,041.59	1.01
Ethane	435.00	20.93	414.06	2.85
Propane	217.77	10.27	207.50	8.29
i-Butane	109.97	4.35	105.62	12.10
n-Butane	109.97	3.47	106.50	16.09
i-Pentane	42.47	1.28	41.18	22.59
n-Pentane	42.47	1.50	40.97	36.94
TOTAL	5,439.92	1,447.05	3,992.87	

CLAIMS:

1. A process for separating a feed gas stream consisting of components which have a spectrum of volatilities including light (volatile) components, intermediate
5 volatility components, and heavy (least volatile) components, comprising the steps of:

(a) contacting the feed gas stream with a lean solvent stream in an absorption zone to produce a light product gas stream comprising light components
10 from the feed gas and a rich solvent liquid stream comprising the lean solvent plus intermediate and heavy components;

(b) flashing the rich solvent liquid stream from the absorption zone in a flashing zone to produce at lower pressure a flash gas stream and a lean solvent
15 stream that is comprised of heavy components that are derived from the feed gas stream; and

(c) conveying the lean solvent produced in step (b) to the absorption zone as the lean solvent stream in step (a).

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2. The process of claim 1 wherein the flash gas stream is comprised of intermediate and heavy components.

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3. The process of claim 2 comprising the additional step of: cooling the feed gas stream in a feed gas cooler to condense a heavy product stream and separating the heavy product stream from the feed gas stream.

30

4. The process of claim 3 comprising the additional step of: adjusting the temperature of the feed gas at the exit of the feed gas cooler to maintain a specified inventory of lean solvent in the flash zone.

5

5. The process of claim 3 comprising the additional step of: conveying heavy product into the flash zone to maintain a specified inventory of lean solvent in the flash zone.

10

6. The process of claim 1 comprising the additional steps of:

(d) compressing and cooling the flash gas stream to form a condensed liquid stream comprising heavy components;

15

(e) separating the condensed heavy component stream from step (d) out of the flash gas stream and conveying it to the flash zone.

20

7. The process of claim 6 comprising the additional step of: adjusting the temperature of the flash gas stream at the exit of the intermediate product cooler to maintain a specified inventory of lean solvent in the flash zone.

25

8. The process of claim 1 comprising the additional steps of:

(d) cooling the lean solvent stream that is conveyed from the flash zone to the absorption zone in step (c) in a lean solvent cooler; and

(e) controlling the temperature of the lean solvent stream leaving the lean solvent cooler to maintain a specified inventory of lean solvent in the flash zone.

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9. The process of claim 1 comprising the additional steps of:

(d) cooling the light product gas stream by direct heat exchange, or by expansion, or by both direct heat exchange and expansion to condense a stream
10 comprised of intermediate and heavy components;

(e) separating the condensate from the light product stream and conveying the condensate into the flash zone.

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10. The process of claim 1 wherein the feed gas is natural gas comprising nitrogen (light component), methane (intermediate component) and hydrocarbons heavier than methane (heavy components) and wherein the light product stream consists essentially of nitrogen and the flash gas stream comprises methane.

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11. The process of claim 10 comprising the additional step of: cooling the feed gas stream to condense a heavy product stream that is comprised of heavy components and separating the condensed stream from the feed gas stream.

25

12. The process of claim 10 wherein the flash gas stream comprises methane and heavy components.

30

13. The process of claim 1 wherein the feed gas is comprised of hydrogen (light component), methane (intermediate component) and hydrocarbons heavier than methane (heavy components) and wherein the light product stream consists essentially of hydrogen and the flash gas stream comprises methane.

5

14. The process of claim 13 comprising the additional step of: cooling the feed gas stream to condense a heavy product stream that is comprised of heavy components and separating the condensed stream from the feed gas stream.

10

15. The process of claim 13 wherein the flash gas stream comprises methane and heavy components.

15

16. The process of claim 1 wherein the feed gas is comprised of hydrogen, nitrogen, methane and wherein the light product stream consists essentially of hydrogen and nitrogen and the flash gas stream comprises methane.

20

17. The process of claim 13 comprising the additional step of: conveying the light product stream to a hydrogen purification unit in which the concentration of hydrogen in the light product stream is increased.

25

18. The process of claim 17 wherein the hydrogen purification unit is a pressure swing adsorption (PSA) unit.

19. The process of claim 16 comprising the additional step of: conveying the light product stream to a hydrogen purification unit in which the concentration of hydrogen in the light product stream is increased.

5

20. The process of claim 19 wherein the hydrogen purification unit is a pressure swing adsorption (PSA) unit.

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1/3

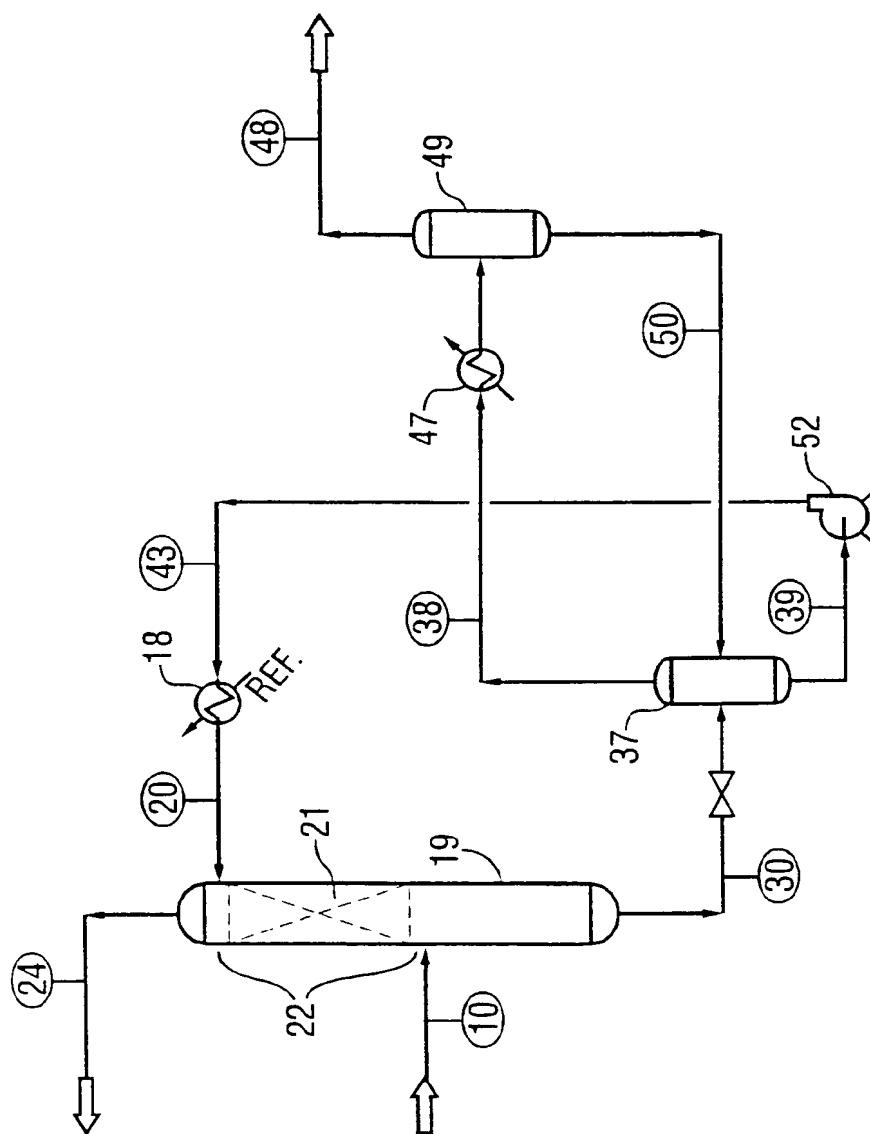
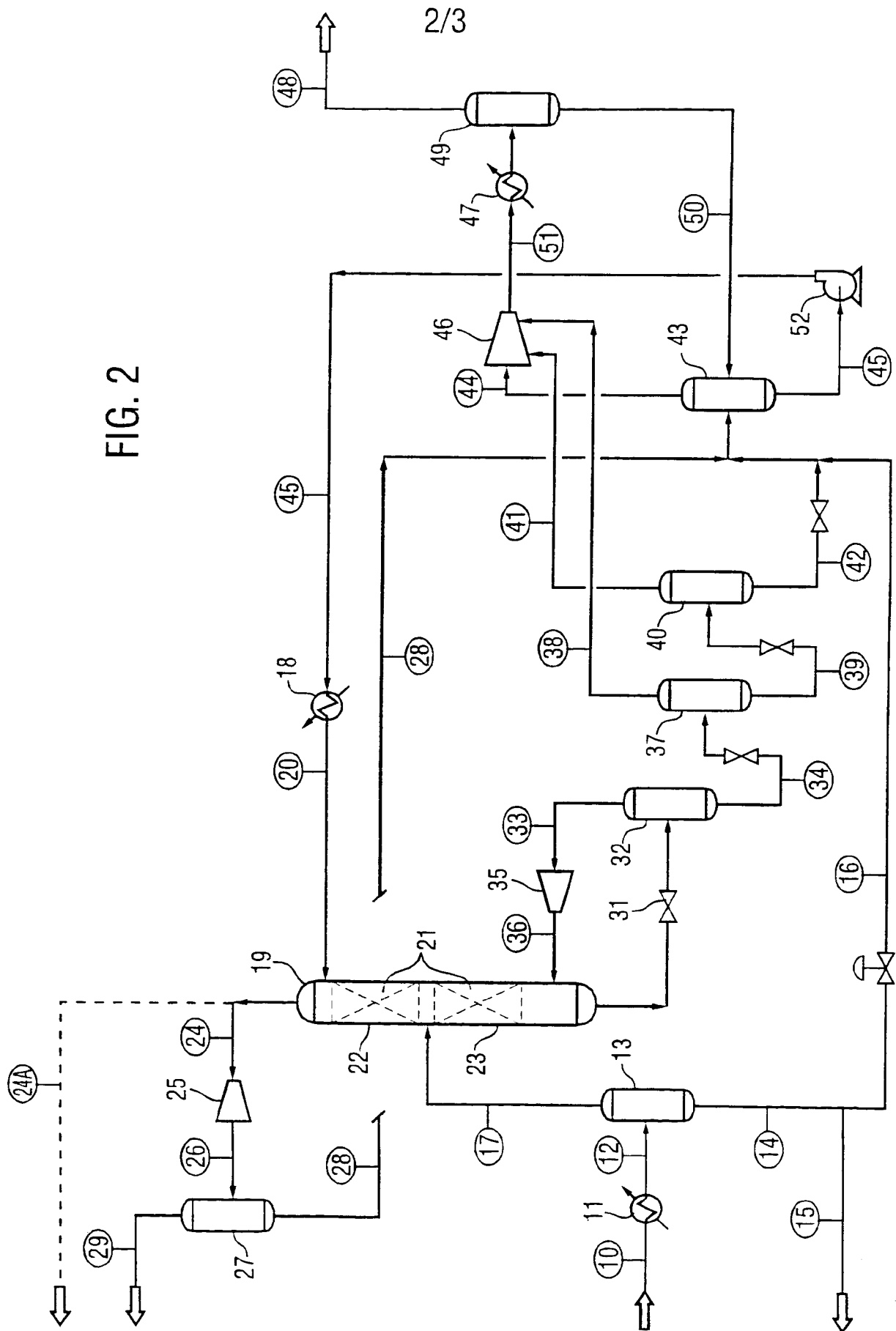
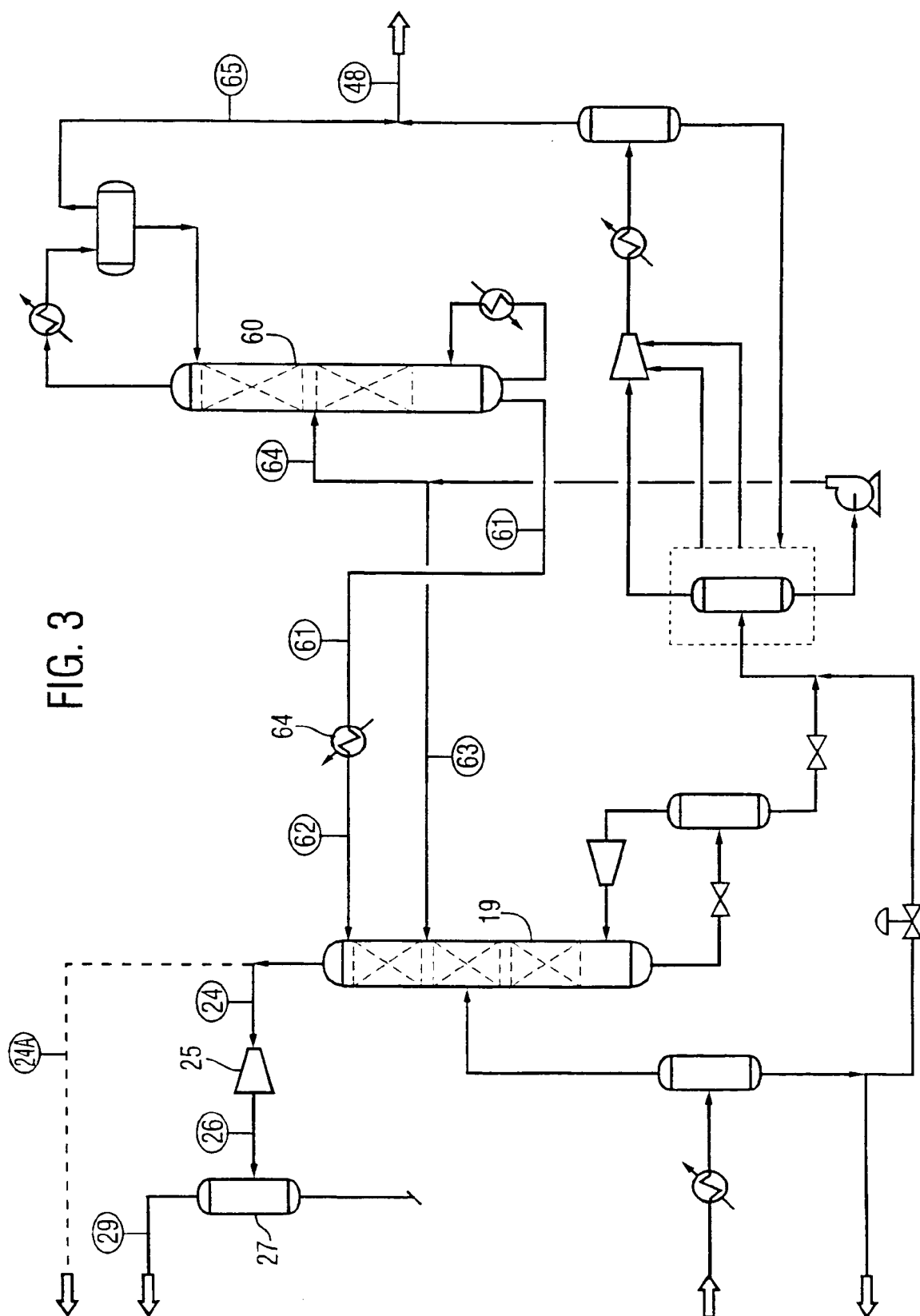


FIG. 1

FIG. 2



3/3



INTERNATIONAL SEARCH REPORT

International Application No.
PCT/US 95/02822

A. CLASSIFICATION OF SUBJECT MATTER

IPC 6 C10G5/04 C01B3/52 C07C7/11 C10L3/10 B01D53/14

According to International Patent Classification (IPC) or to both national classification and IPC.

B. FIELDS SEARCHED

Minimum documentation searched (classification system followed by classification symbols)

IPC 6 C10G C01B C07C C10L B01D

Documentation searched other than minimum documentation to the extent that such documents are included in the fields searched

Electronic data base consulted during the international search (name of data base and, where practical, search terms used)

C. DOCUMENTS CONSIDERED TO BE RELEVANT

Category *	Citation of document, with indication, where appropriate, of the relevant passages	Relevant to claim No.
P,X	US,A,5 321 952 (UOP) 21 June 1994 see column 5, line 4 - line 68 see column 7, line 52 - column 8, line 2 see column 9, line 57 - column 10, line 13; figures 1-3 ---	1-8, 10-16
X	DE,A,31 32 292 (LOHMANN GMBH & CO. KG) 3 March 1983 see page 6, paragraph 4 - page 7, paragraph 1; claims 1-13; figure 1 ---	1
X	US,A,4 368 058 (PHILLIPS PETROLEUM CO.) 11 January 1983 see column 2, line 8 - line 13; claims 1-6; figure 1 see column 1, line 34 - line 39 --- -/--	1



Further documents are listed in the continuation of box C.



Patent family members are listed in annex.

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- *A* document defining the general state of the art which is not considered to be of particular relevance
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- * & * document member of the same patent family

Date of the actual completion of the international search

21 June 1995

Date of mailing of the international search report

30.06.95

Name and mailing address of the ISA

European Patent Office, P.B. 5818 Patentlaan 2
NL - 2280 HV Rijswijk
Tel. (+ 31-70) 340-2040, Tx. 31 651 epo nl,
Fax: (+ 31-70) 340-3016

Authorized officer

Eijkenboom, A

INTERNATIONAL SEARCH REPORT

International Application No.
PCT/US 95/02822

C.(Continuation) DOCUMENTS CONSIDERED TO BE RELEVANT

Category *	Citation of document, with indication, where appropriate, of the relevant passages	Relevant to claim No.
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X	US,A,4 333 819 (UOP INC.) 8 June 1982 see column 5, line 32 - column 6, line 14; figure 1 ---	1-5,8, 13-15
X	DE,A,30 31 727 (LINDE AG) 1 April 1982 see page 9, line 25 - page 10, line 4; claims 1-12; figure 1 see page 7, line 6 - page 8, line 15 see page 6, line 12 - line 25 -----	1-8, 13-16

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International Application No.

PCT/US 95/02822

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